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OF MODERN SOCIETY

Flue Gas Desulfurization Pilot Study Phase I — Survey of Major Installations - Double Alkali/Gypsum Flue Gas Desulfurization Process

Appendix 95-G

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16. ABSTRACT The double alkali/gypsum of flue gas desulfurization (FGD) was developed to reduce scaling and plugging problems associated with direct lime (CaO) or limestone ($CaCO_3$) systems. One of the first large-scale installations of this process to control SO_2 emissions and produce gypsum ($CaSO_4$) was put in operation in 1973 at the Showa Denko plant in Chiba, Japan. By December 1977, 25 systems were operating, all in Japan. The double alkali system is especially applicable in situations requiring high (98 percent) SO_2 removal, high availability, and low maintenance. The process is particularly applicable to areas where lime/limestone resources are abundant and where sypsum is a marketable by-product. If it is not marketable, gypsum and fly ash can be used as landfill to form a base with load-bearing capacities sufficient for some construction.

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Flue Gas Desulfurization Pilot Study Phase I

Survey of Major Installations - Double Alkali/ Gypsum Flue Gas Desulfurization Process

APPENDIX 95-G

Lead Country - United States

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SUMMARY

The double alkali/gypsum system of flue gas desulfurization (FGD) was developed to reduce scaling and plugging problems associated with direct lime (CaO) or limestone ($CaCO_3$) systems. One of the first large-scale installations of this process to control SO_2 emissions and produce gypsum ($CaSO_4$) was put in operation in 1973 at the Showa Denko plant in Chiba, Japan. By December 1977, 25 systems were operating, all in Japan. The double alkali system is especially applicable in situations requiring high (98 percent) SO_2 removal, high availability, and low maintenance. The process is particularly applicable to areas where lime/limestone resources are abundant and where gypsum is a marketable by-product. If it is not marketable, gypsum and fly ash can be used as landfill to form a base with load-bearing capacities sufficient for some construction.

The system uses a soluble alkali salt as the absorbing solution. After adsorbing SO_2 in a gas-liquid contactor, the solution forms a soluble reaction product. Potassium, ammonium, or aluminum salt solutions can be used in such systems, but a sodium solution is the most popular choice.

The absorbent is then reacted with lime or limestone to form an insoluble product comprising calcium sulfite hemihydrate ($CaSO_3 \cdot 1/2\ H_2O$), gypsum ($CaSO_4 \cdot 2H_2O$), and the original alkali salt. The solids are further oxidized to produce more gypsum. Some systems oxidize the scrubber absorbant first, then regenerate the oxidized reaction products with lime or limestone. The gypsum sludge is washed to remove the remaining alkali and then dewatered. Gypsum produced in this manner is of a higher quality than that produced by a direct lime/limestone absorption system and is suitable for either cement or wallboard.

The advantages and disadvantages of the double alkali/gypsum system are summarized below:

Advantages

1. The soluble product in the absorber minimizes solids build-up and erosion problems, thereby offering high system availability.
2. Very high SO_2 removal efficiencies are obtained.
3. Because particulates are removed in the quenching and scrubbing stage, the process can simultaneously remove SO_2 and particulates unless gypsum is to be sold for wallboard manufacture.
4. Absorption takes place in the scrubber in an unsaturated solution: therefore, the absorbant has a large absorption capacity.
5. A relatively low liquid-to-gas (L/G) ratio is required in the scrubber.
6. A salable by-product (gypsum) is produced.

7. When no market is available, gypsum can be used as a landfill and provides a good stable construction base compared with calcium sulfite sludge.

8. Corrosion and erosion problems are minor compared with those in wet lime/limestone processes.

Disadvantages/Problems

1. Design complexities must be introduced to circumvent the following problems:

   (a) Excessive purge of nonregenerable Na_2SO_4, produced as a result of oxidation.

   (b) Scaling problems resulting from scrubbing with a clear liquor saturated with calcium sulfate.

2. Some makeup is required to replenish sodium losses.

3. Installation costs are high compared with those of the lime/limestone gypsum system.

1.0 BACKGROUND AND INTRODUCTION

This report is one of a series on the status of flue gas desulfurization (FGD) processes in the NATO countries and Japan. The United States, the Federal Republic of Germany, and Norway acted as lead countries in gathering information and preparing reports on selected FGD systems.

The double alkali/gypsum process is an indirect lime/limestone process using a soluble regenerable sodium, potassium, ammonium, or aluminum salt solution as the absorbant solution. The SO_2 is adsorbed by the solution and then reacted with a lime/limestone slurry in a separate reactor. The sludge produced is oxidized to gypsum, a salable by-product. The double alkali process was developed primarily to reduce the scaling and plugging problems associated with direct lime/limestone processes. To date, Japan is the only country where the marketability of gypsum and the limited land disposal sites are such that the production of gypsum can be justified.

One of the first large-scale installations of the system was on a boiler at the Showa Denko plant in Chiba, Japan. This system was put in operation in January 1973

with a capacity equivalent to 150 MW. Limestone is used to regenerate the sodium sulfite (Na_2SO_3) scrubbing solution. The system produces about 4000 kg/h of large-grain, high-quality gypsum. In 1975, nine industrial boilers and one 150-MW utility boiler in Japan started using double alkali/gypsum systems to control SO_2. Twenty-five such systems (including seven utility boilers representing a capacity of 1949 MW) were in operation by December 1977.

As operation of wet lime/limestone processes becomes more successful, double alkali processes are losing their advantage. Sodium-based indirect processes still may be used, however, where very high SO_2 removal is required, even though they are more expensive than lime/limestone systems. Double alkali processes also are especially useful in preventing scaling at plants where operation cannot be readily controlled because of sulfur or flow variations.

2.0 GENERAL PROCESS DESCRIPTION

The double alkali technique for controlling SO_2 emissions involves absorbing the SO_2 by a soluble alkaline medium in the scrubbing chamber and treating the scrubber effluent with lime or limestone in a separate reactor outside the scrubber loop. Thus, a double alkali FGD system is an indirect lime/limestone scrubbing method that removes SO_2 without the potential for plugging and scaling that is found in direct lime/limestone scrubbing systems. Calcium sulfites and sulfates are precipitated, while the scrubbing liquor is regenerated and returned to the scrubbing loop. The precipitated sludge is oxidized to gypsum, a salable by-product. Figure 1 shows an example double alkali FGD system using sodium as the absorbant, as applied to a 500-MW coal-fired boiler.

2.1 PROCESS CHEMISTRY

2.1.1 Absorption

Sulfur dioxide removal is effected in the gas scrubber by the reaction of the gas with a circulating alkaline solution, which is sprayed countercurrent to the gas flow. Although the scrubbing liquor is generally a sodium solution

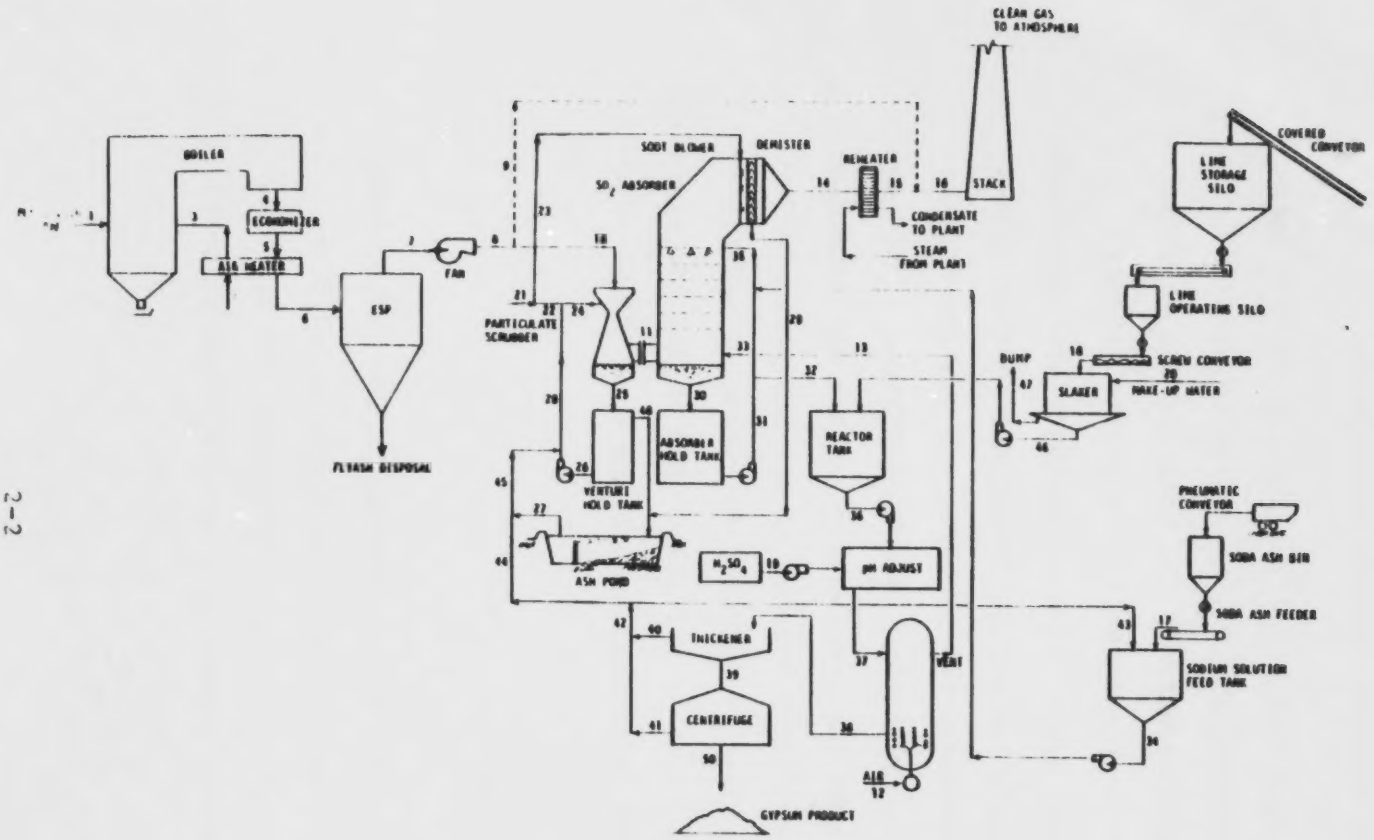

Figure 1. Example double alkali based, nonregenerable FGD system (Gypsum production). Flow sheet and material balance applied to a 500-MW, coal-fired boiler. (See notes at end of figure.)

(continued)

Stream No.	1	2	3	4	5	6	7	8	9	10	11
Description	Coal to boiler	Air to heater	Air to boiler	Gas to economizer	Gas to air heater	Gas to ESP	Gas to fan	Fan outlet gas	Gas bypass	Gas to venturi	Gas to absorber
Rate, 1000 kg/h	170	2049	1848	1998	1998	2199	2186	2186	77.3	2109	2191.6
Nm^3/s		464	419	445	445	490	490	490	17	473	503
m^3/s liquid											
Particulates, kg/h				15,300	15,300	15,300	2600	2600	91	2510	12.2
Temperature, °C		43.3	279	477	374	154					51.7
Specific gravity											
Viscosity, Pascal-seconds											
Undissolved solids, %											
pH											
Sulfur as SO_2 kg/h	11,900			11,900	11,900	11,900	11,900	11,900	420	11,480	10,330

Stream No.	12	13	14	15	16	17	18	19	20	21	22
Description	Air to oxidizer	Air vent to absorber	Gas to reheater	Gas from reheater	Gas to stack	Soda ash	Lime to slaker	H_2SO_4 for pH adjust	Water for slaking	Makeup water	Water to venturi
Rate, 1000 kg/h	61.6	62.7	2244.1	2244.1	2321.4	0.79	11.46	21.1	62	18.8	11.2
Nm^3/s	13.3	13.8	516.4	516.4	533.4						
m^3/s liquid									0.0172	0.0052	0.0031
Particulates, kg/h			12.2	12.2	103.2						
Temperature, °C	20	35	51.7	78.9	78.9						
Specific gravity									1.0	1.0	1.0
Viscosity, Pascal-seconds											
Undissolved solid, %								0	0	0	0
pH											
Sulfur as SO_2 kg/h			770	770	1190						

(continued)

Figure 1 (continued).

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Stream No.	23	24	25	26	27	28	29	30	31	32
Description	Demister cleaning water	Slurry to venturi	Spent slurry from venturi	Slurry recycle from venturi	Pond water recycle	Recycle slurry to venturi	Spent water from demister	Spent scrubbing solution	Scrubbing solution recycle	Purge to reactor
Rate, 1000 kg/h	7.6	4597.4	4515	4465	45.4	4586.2	7.6	5029.6	5037.2	110.2
Nm^3/s										
m^3/s liquid	0.0021	1.240	1.218	1.204	0.0126	1.237	0.0021	1.397	1.399	0.0306
Particulates, kg/h		223,250	225,750	223,250		223,250				
Temperature, °C										
Specific gravity	1.0	1.03	1.03	1.03	1.0	1.03	1.0	1.0	1.0	1.0
Viscosity, Pascal- seconds										
Undissolved solids, %	0	5	5	5	0	5	0	0	0	0
pH										
Sulfur as SO_2 kg/h		102,520	101,700	102,520		102,520				

Stream No.	33	34	35	36	37	38	39	40	41
Description	Recycle less purge	Sodium solution	Scrubbing solution to absorber	Reactor underflow	Oxidizer feed	Thickener feed	Thickener underflow	Thickener overflow	Concentrate
Rate, 1000 kg/h	4927	92.4	5019.4	183.3	204.4	203.1	101.7	101.6	65.8
Nm^3/s									
m^3/s liquid	1.369	0.0257	1.394	0.0467	0.0526	0.0518	0.0235	0.0282	0.0183
Particulates, kg/h									
Temperature, °C									
Specific gravity	1.0	1.0	1.0	1.09	1.08	1.09	1.2	1.0	1.0
Viscosity, Pascal- seconds									
Undissolved solids, %	0	0	0	15	13.5	15	30	0	0
pH									
Sulfur as SO_2 kg/h				9560	9560	9560	9560		

(continued)

Figure 1 (continued).

Stream No. / Description	42 Recovered water	43 Water to sodium feed tank	44 Makeup venturi system	45 Recycle slurry	46 Slurry to reactor	47 Grit dump	48 Particulate slurry product	49 Fly ash disposal	50 Gypsum product
Rate, 1000 kg/h	167.4	91.6	75.8	121.2	73.13	0.34	50	12.7	35.9
Nm^3/s									
m^3/s liquid	0.0465	0.0254	0.021	0.0337	0.0185		0.0135		
Particulates, kg/h							2500	12,700	
Temperature, °C									
Specific gravity	1.0	1.0	1.0	1.0	1.1		1.03		
Viscosity, Pascal-seconds									
Undissolved solids, %	0	0	0	0	20		5		85
pH					12.1				
Sulfur as SO_2 kg/h							1150		9560

Notes:

1. Calculations based on:

   a. 130% stoichiometric lime; 95% CaO and 5% grit.
   b. 60% of the grit is removed.
   c. 3.5% sulfur in coal (dry).
   d. 12% ash coal (as fired).
   e. 92% sulfur in coal evolves as SO_2.
   f. 75% of ash in coal evolves as fly ash.
   g. 99.5% removal of fly ash to scrubber.
   h. 80% solids in ash pond.
   i. Sulfate-to-sulfite ratio in feed to oxidizer is 6:4.
   j. 5% SO_2 removed in soda ash lost.

Figure 1 (continued).

made up of sodium hydroxide (NaOH) and sodium sulfite ($NaSO_3$), it can be alkali of potassium, ammonium, or aluminum.

The scrubber liquor is usually defined as either concentrated or dilute with respect to NaOH concentration. Solutions of greater than 0.1 molal of NaOH are considered concentrated. The choice of liquor is based strictly on the oxidation potential of the flue gas. Concentrated systems are better for units with low oxidation rates, such as boilers burning high-sulfur coal or large boilers on which excess air is carefully controlled. Dilute systems are better for small boilers, boilers burning low-sulfur coal, or boilers with large amounts of excess air. The concentrated system, which is the predominant choice, is described herein.

The sulfur dioxide is absorbed by the solution and produces a spent sodium sulfite/bisulfite liquor according to the following reaction sequence:

$$NaOH + SO_2 \rightarrow NaHSO_3 \quad (1)$$

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \quad (2)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \quad (3)$$

In reaction 1, SO_2 is absorbed by the sodium hydroxide to form bisulfite. The same reactants also unite, as shown in reaction 2, to form sodium sulfite, which further absorbs SO_2 to form bisulfite in reaction 3. The scrubber liquor is pumped to the lime or limestone regenerator.

2.1.2 Regeneration of Scrubbing Solution

The SO_2 removed from the flue gas in the form of soluble sodium sulfite-bisulfite is reacted with a lime or limestone slurry. This transfers the SO_2 to the lime/limestone slurry, where it reacts to form insoluble calcium sulfite and calcium sulfate. This regenerates the sodium sulfite, which is returned to the scrubber after it is softened, and additional sodium salts are added. (Any chemical reaction that removes excess minerals from solution to reduce scaling potential is known as a "softening" reaction.)

When limestone (calcium carbonate) is used, regeneration takes place in several stages as shown in the following series of reactions:

$$NaHSO_3 + CaCO_3 \rightarrow NaHCO_3 + CaSO_3 \qquad (4)$$

Because sodium bisulfite is by far the major reactant, the solution remains acidic. In the acid solution, the sodium bicarbonate will break down to sodium hydroxide and carbon dioxide.

$$NaHCO_3 \rightarrow NaOH + CO_2 \qquad (5)$$

The sodium hydroxide then reacts with the excess sodium bisulfite to produce additional scrubbing liquor.

$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O \qquad (6)$$

The overall regeneration reaction is, therefore:

$$2NaHSO_3 + CaCO_3 \rightarrow Na_2SO_3 + CaSO_3 + H_2O + CO_2 \quad (7)$$

When lime is used for regeneration, it must first be slaked or hydrated.

$$CaO + H_2O \rightarrow Ca(OH)_2 \quad (8)$$

The slaked lime is then pumped to the reactor tank, where the following overall regeneration occurs:

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 + 2H_2O \quad (9)$$

The lime further reacts with the sodium sulfite and sodium sulfate in solution. These reactions are limited by the concentration of the equilibrium hydroxide ions and the sulfite ions, which are more active than the sulfate species when forming sodium hydroxide.

$$Na_2SO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_3 \quad (10)$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4 \quad (11)$$

2.1.3 Oxidation of Sulfites to Sulfates

The regeneration reaction forms both sulfites and sulfates. The sulfites must be oxidized to sulfates to form a pure gypsum by-product. Oxidation of sulfites to sulfates is carefully controlled. Unrestricted oxidation in a concentrate-mode scrubber results in scaling and loss of active scrubbing species (Na_2SO_3) to the inactive species (Na_2SO_4). Even in carefully controlled systems such as those in utility power plants, oxidation of Na_2SO_3 to Na_2SO_4 can be as high as 5 to 10 percent of the SO_2 removed. After the

scrubber liquor is reacted with SO_2, oxidation of sulfites to sulfates is encouraged. Sulfites can be converted to sulfates at two points in the system, the end product being gypsum. The Na_2SO_3 scrubbing liquor is contacted with air to oxidize $SO_3^=$ into $SO_4^=$; then lime or limestone is added in the reactor to precipitate gypsum. An alternative is to react the Na_2SO_3 scrubbing liquor with lime or limestone first and then contact the $CaSO_3$ with air to produce gypsum.

The slurry is pumped from the reactor tank to a clarifier, or thickener, where the precipitated solids are separated from the scrubbing liquor. The solids are washed of residual sodium and dewatered. Finally the by-product gypsum is stockpiled.

2.1.4 Sodium Makeup and Softening

The liquor from the clarifier is returned to the scrubbing loop, but not before it has been treated to correct two potential problems. First, makeup sodium or other alkali must be added to the solution, because a small percentage of the scrubbing liquor is lost in the gypsum product. Second, any excess calcium ion concentration must be reduced to prevent scaling in the scrubber. Both calcium sulfate and calcium sulfite are produced during regeneration, as shown in reactions 10 and 11. The solubility of calcium sulfate is approximately two orders of magnitude greater

than that of calcium sulfite; therefore, calcium sulfate is more likely to remain in solution and be carried in the liquor from the clarifier. In other words, when solid calcium sulfate is present in solution at equilibrium, the regeneration system produces high calcium ion concentrations. As conditions develop that cause the scrubbing liquor to become supersaturated with calcium ions, a high potential for scaling occurs in the scrubber.

If limestone regeneration is used, lost sodium is replaced by adding NaOH to the solution. Only as much sodium hydroxide is added to the liquor as is necessary to make up for sodium lost in the waste discharge. The excess calcium can be removed by contacting the solution with a stream of clean boiler flue gas containing carbon dioxide. The carbon dioxide reacts with the calcium ions as presented in the following equation.

$$Ca^{++} + CO_2 + H_2O \rightarrow CaCO_3 + 2H^+ \qquad (12)$$

The calcium ion concentration is now reduced to a level below the saturation value, and the liquor can be safely returned to the scrubbing loop. The precipitated calcium carbonate settles out and can be used to regenerate the scrubbing liquor as shown in reaction 7.

The same procedure for replenishing lost sodium and reducing calcium ion concentrations in the liquor may be

used in a lime regeneration system. When using lime, however, an alternate method not applicable to a limestone regeneration system is preferred. In this method, sodium carbonate (Na_2CO_3) is added to the liquor from the clarifier and not only replaces lost sodium but also softens the solution in one chemical reaction.

$$Na_2CO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaCO_3 \qquad (13)$$

Sodium is replaced in the scrubbing liquor as sodium hydroxide, and excess calcium is removed as precipitated calcium carbonate. As in the limestone system, only enough sodium is added to the liquor to make up for that lost in the waste discharge. The amount of sodium carbonate added may be insufficient to soften the solution satisfactorily, however; and CO_2 may be added (as in reaction 12) to complete the softening process.

The makeup sodium carbonate requirement is estimated at 5 percent on a weight basis and can be calculated according to equation 4 and molecular weights:

$$\text{Makeup soda ash required} = \frac{106}{64} \times \frac{5}{100} = 0.083 \text{ kg/kg of } SO_2.$$

Lime is required in the reactor to regenerate sodium solutions. According to equations 8, 9, 10, and 11, theoretical lime requirement = $\frac{56}{64}$ = 0.875 kg/kg of SO_2. Using 1.1 of stoichiometric requirement and 95 percent lime purity, the actual lime requirement = $\frac{56}{64} \times \frac{1.1}{0.95}$ = 1.0132 kg/kg of SO_2.

The gypsum generated at the reactor is similar to that generated in a lime or limestone process but tends to be of a quality more suitable for wallboard. The unused lime, lime impurities, and make-up sodium carbonate also combine with the gypsum.

2.1.5 Alternative Operating Modes

The various alternative steps described above can be combined in FGD systems in a number of ways. The five basic modes of sodium/calcium scrubbing are defined below:

1. Limestone and lime regeneration, dilute active alkali, with sulfite softening.

2. Lime regeneration, dilute active alkali, with sulfite softening.

3. Lime regeneration, concentrated active alkali, with side-stream sulfate treatment (removal).

4. Limestone regeneration, concentrated active alkali, with side-stream sulfate treatment (removal).

5. Lime regeneration, dilute active alkali, with carbonate softening.

These modes are discussed briefly in the following sections, and some advantages and disadvantages of each are noted. Many variations of these five basic schemes can be produced by varying certain parameters associated with the systems.

The first mode identified is a double-loop (limestone/lime), dilute alkali system. Dilute alkali indicates that

the sulfite or alkali concentration entering the scrubber will be less than approximately 0.1 molality. Dilute alkali scrubbing solutions allow sulfate regeneration with lime in this mode of operation. "Double loop" refers to the regeneration method. Spent scrubbing liquor is treated with limestone to neutralize the bisulfite and precipitate calcium sulfite. A stream from this limestone reaction vessel is then treated with lime to precipitate calcium sulfite and sulfate and generate sodium hydroxide.

The relative amount of reaction between lime and the sulfite and sulfate is determined mainly by sulfite and sulfate concentrations and lime stoichiometry. Liquor from the lime reaction vessel is returned to the limestone reaction vessel to ensure that unsaturated calcium ion concentrations go to the scrubber. Desaturation is accomplished by the presence of sulfite ion in the limestone reactor. The reduction of calcium ion concentrations by calcium sulfite precipitation is referred to as sulfite softening. Sodium carbonate may be added to provide additional softening and make up any sodium losses.

The disadvantages of this scheme are that it requires relatively long hold times for the reaction with limestone, it requires two major reactors, and it must circulate large volumes of relatively dilute scrubbing liquor. The ad-

vantages are that the limestone used for regeneration of both bisulfite and sulfate is relatively inexpensive and equipment is not complex.

The second mode of operation is a double-loop (lime only), dilute alkali system. The arrangement is very similar to that of the first mode. Spent scrubbing liquor is treated with lime to neutralize the bisulfite only. A stream from the lime/bisulfite reaction vessel is then treated with additional lime in a separate reaction vessel to react with sulfite and sulfate to form calcium solids and sodium hydroxide. Liquor from this second reaction vessel is returned to the lime/bisulfite reaction vessel to ensure unsaturated calcium ion concentrations with respect to calcium sulfate by sulfite softening of the liquors before they enter the scrubber.

The disadvantages of this flow scheme are the difficulty of controlling pH and calcium ion control in the regeneration loop [because of the addition of a relatively concentrated strong base (lime) to a dilute solution of a weak acid], the necessity of two reactors for the regeneration loop, and the need to circulate large volumes of relatively dilute scrubbing liquor to the scrubber. The advantages are the relatively short hold times necessary for regeneration with lime and the fairly simple equipment requirements.

The third mode of operation involves lime regeneration, concentrated alkali solutions, and independent treatment of sulfate removal and control. Lime is used for the bisulfite neutralization of the concentrated (∿0.5 molality) alkali scrubbing solutions. A side stream is taken from the bisulfite reactor, where sulfuric acid is used to precipitate sulfates and regenerate the scrubbing liquor. Again, sulfite softening is used to remove residual calcium ions. Figure 1 shows a typical flow sheet for this mode.

This third mode has the disadvantage of potential sulfate regeneration complications. The advantages of the scheme are that relatively short hold times are necessary for bisulfite/sulfite regeneration with lime, utilization of reactants is high, the state of the art is advanced, and high SO_2 removal efficiencies are achieved with low flows of the concentrated alkali scrubbing liquors. This third mode of operation appears to be very promising, but universal, large-scale, successful application depends on how effective the treatment is for sulfate removal and control.

The fourth mode involves limestone regeneration, concentrated active alkali, and side-stream sulfate removal. This mode differs from the third mode only in that limestone is used instead of lime for the bisulfite-removal step. Like the third mode, it has the possible disadvantage of

sulfate regeneration complications, and it has the added disadvantage that limestone requires long hold times for the bisulfite/sulfite regeneration. Advantages of this flow scheme are that SO_2 removal efficiencies and utilization of reactants are high, and liquor flow rates are low. Successful large-scale application of this flow scheme also depends on how effective the treatment is for sulfate removal and control.

The fifth mode of operation is a lime regeneration, dilute alkali system with carbonate softening for calcium ion (scaling) control. The spent scrubbing liquor is treated with lime, which neutralizes bisulfite and reacts with sulfite and sulfate. The only significant difference between this process and the second mode is the use of carbonate softening (CO_2 and Na_2CO_3) to lower the calcium ion concentration. The disadvantages of this scheme are that it may not regenerate sulfate adequately, it has the potential for scaling, and it must circulate large volumes of relatively dilute scrubbing liquors. The advantages are that it can deal with high oxidation levels, its equipment requirements are relatively simple, and the state of the art is advanced.

2.2 DESCRIPTION OF EQUIPMENT COMPONENTS

2.2.1 Particulate Removal

Before sulfur oxides are removed, flue gas is passed through a venturi scrubber or an ESP, which removes a large amount of any particulate matter that may be in the gas stream. This not only prevents the emission of particulates, but also reduces the potential for plugging in the gas scrubbing unit and reduces contamination of the scrubbing liquor and gypsum. Particulate scrubbing liquor is kept separate from SO_2 scrubbing liquor. The spent particulate scrubbing liquor is pumped to an ash pond, where the fly ash settles and the water is neutralized with lime or limestone; it then is returned to the venturi. Any dry fly ash from an ESP is combined with the venturi liquor before disposal.

2.2.2 SO_2 Absorber

The absorber is the primary SO_2 removal unit in the double alkali/gypsum FGD system. Each of the many designs has a different method of contacting the flue gas with the slurry. The more common unit designs are spray towers (Figure 2), packed towers (Figure 3), venturi scrubbers (Figure 4), and sieve or perforated trays (Figure 5).

As with calcium-based systems, corrosion and pitting can be a problem. Selection of proper materials for the scrubber body and internals is necessary for long life and

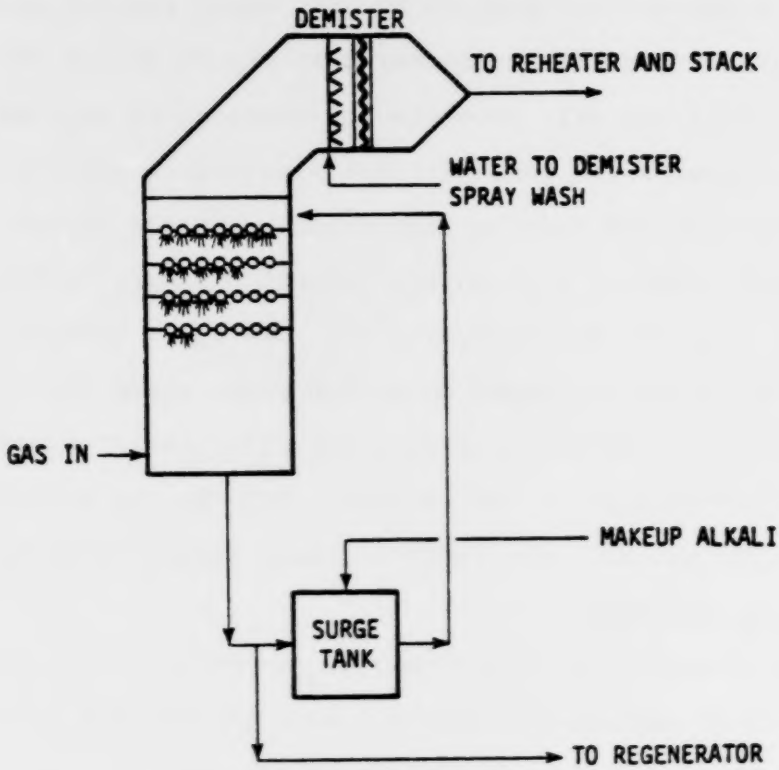

Figure 2. Spray tower.

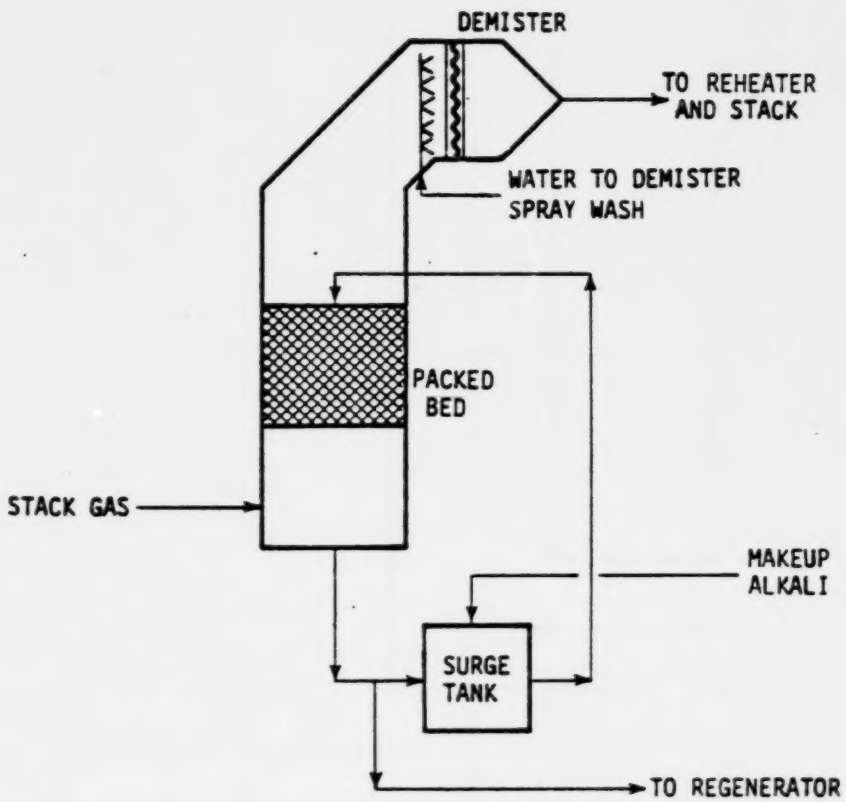

Figure 3. Fixed packing absorber.

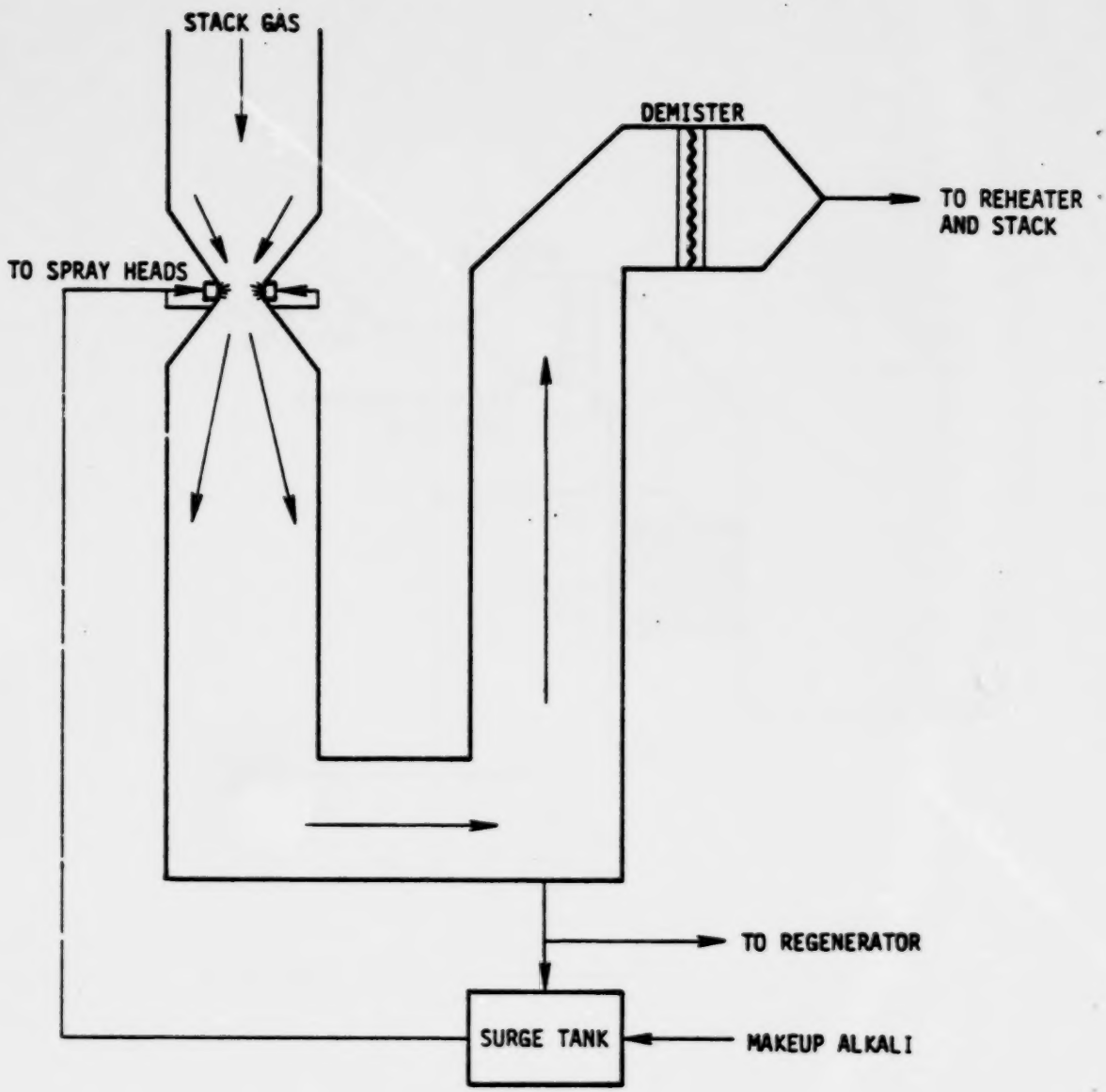

Figure 4. Venturi type absorber.

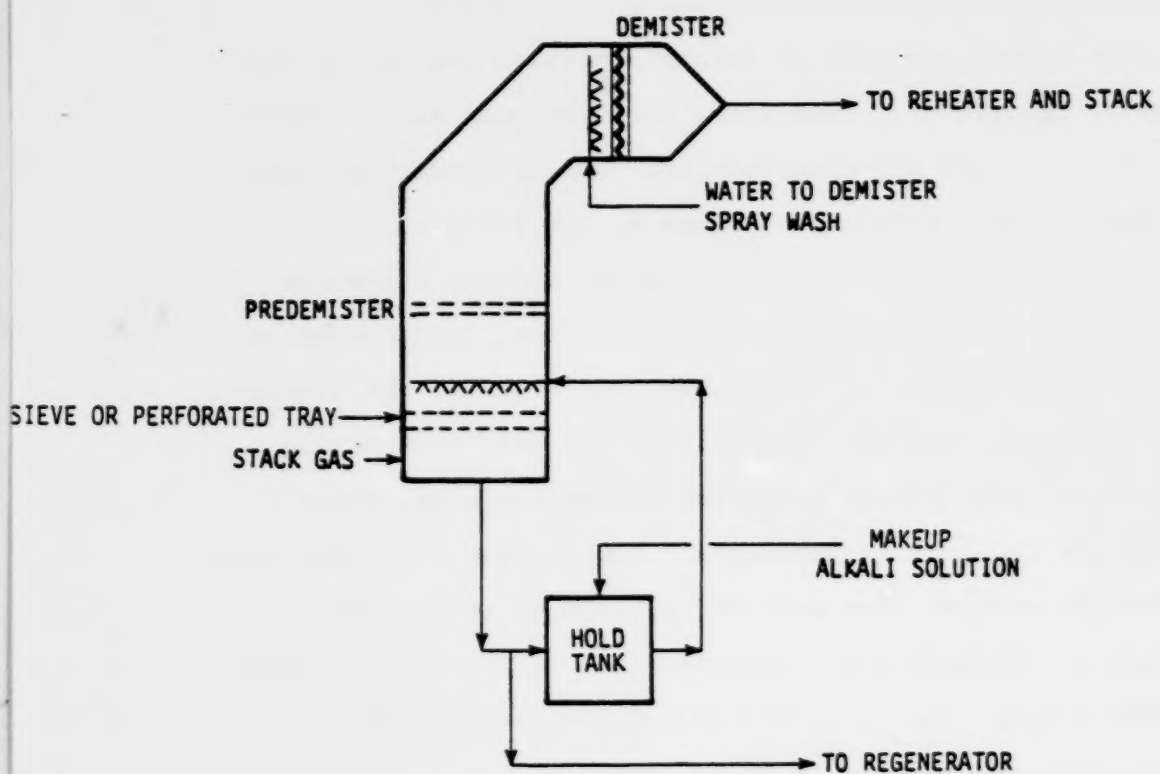

Figure 5. Sieve or perforated tray absorber.

maintenance-free operation. Materials that have been used for the scrubber body include: steel lined with rubber, fiberglass-reinforced polyester (FRP), acid brick, or flaked glass. Sometimes 316L stainless steel is used. Materials that have been used for absorber internals include FRP, 316 stainless steel, polyethelene, and polypropylene.

The number and size of scrubber modules per system are directly related to boiler load, turndown potential, system availability, and L/G distribution. Boiler loads fluctuate, and the scrubber system must adapt to low loads without reduced performance. Multiple scrubber modules allow the shutdown of individual modules as the boiler load decreases, and the more modules in the system, the smoother the transition. Because scrubber modules that are shut down during low boiler loads can be scheduled for maintenance, overall system availability is increased. The smaller cross-sections of smaller modules also make it possible to achieve more uniform L/G distribution. Absorber module sizes of surveyed systems ranged from 43 to 225 MW and averaged 103 MW.

2.2.3 Demister

The demister downstream from the absorber is necessary to remove entrained water droplets, thereby decreasing downstream equipment corrosion and decreasing energy requirements in the reheater. Most droplets are generally

large enough to be removed with a simple change in gas direction. Mesh mist pads or chevron type cyclonic vane demisters are sufficient, and they can be placed in series for additional capacity. Demisters are subject to corrosion, necessitating the installation of a soot blower or spray wash. Demisters are constructed of either stainless steel or fiberglass-reinforced polyester, and can be installed in the horizontal or vertical position. The vertical position (as shown in Figures 1 to 5) is more popular in Japan.

2.2.4 Stack Gas Reheater

It is generally necessary to reheat stack gas to maintain plume dispersion. Some secondary advantages of reheating are that it protects downstream equipment (such as ducts and stacks) and prevents acid mist formation. Reheating can be accomplished by installing a gas or oil burner that exhausts directly into the stack, installing steam coils to heat flue gas, using steam coils to heat air and injecting it into the stack, or bypassing flue gas around the FGD unit into the stack. An in-line heat exchanger costs the least initially, but maintenance costs are high because it is subject to corrosion. A direct-fired afterburner is used most often to reheat flue gas in the double alkali/gypsum system.

2.2.5 Regeneration

Regeneration occurs in a continuously stirred tank reactor. The size of the reactor depends on the process, whether a dilute or concentrated mode is used, and whether lime or limestone regeneration is applied. Reaction hold tank time is about ten times longer for limestone than for lime. Materials of construction for reaction tanks are similar to those of the absorber.

2.2.6 Oxidation

Calcium sulfite sludge from the regenerator is first adjusted to a pH of 3 to 4 by adding sulfuric acid, then pumped to the oxidation tower. Air is pumped into the bottom of the oxidation tower, and the sludge flows concurrently down from the top. The oxidizer is operated at above atmospheric pressure usually at 200 to 400 kPa. Air from the top of the tower is vented to the scrubber to remove any SO_2 volatized in the oxidizer. The sludge of calcium sulfate slurry and sodium sulfite is pumped to a centrifuge for separation and washing. Materials of construction of the oxidation tower are similar to those in the absorber.

2.2.7 Centrifuge

The centrifuge separates the solid gypsum ($CaSO_4 \cdot 2H_2O$) from the liquid solution of Na_2SO_3, $NaHSO_3$ and Na_2SO_4.

After separation, the solid cake is washed to remove additional sodium and any other soluble impurities. The by-product gypsum, usually 6 to 15 percent water, is stored until it can be sold or used as landfill. The gypsum crystals produced by this process can be as large as 0.07 mm thick and 0.5 mm long. They are excellent for wallboard and can easily be dewatered down to 5 percent water. A basket centrifuge is generally used, but other types are acceptable. A vacuum filter can also be used. The materials of construction are similar to those of the absorber.

2.3 OPERATING PARAMETERS

2.3.1 Chemical Parameters

In a double alkali system, the pH of the absorber is maintained in the range of 6 to 7, but 6.5 is the normal control point. A higher pH increases carbon dioxide absorption and can lead to carbonate scaling, whereas a lower pH increases the SO_2 vapor pressure over the solution and reduces SO_2 absorption. A solution pH of 6.5 is highly buffered and can adapt rapidly to flue gas inlet changes. A more detailed description of the above phenomenon is presented in Section 4.0

The solution from the absorber is pumped to the regeneration reactor. Limestone is added and the pH increases to about 7.3. After regeneration, the pH is reduced to 3 or

4 by adding sulfuric acid and pumped to the oxidation tower. The low pH promotes oxidation of $CaSO_3$ to $CaSO_4$.

2.3.2 Physical Parameters

The primary scrubber in the system can be a particulate scrubber or just a gas cooler and humidifier. Flue gas temperature is normally lowered to 55°C to provide adequate SO_2 removal. Liquid to gas ratios of the surveyed systems range from 0.5 to 0.6 liter/Nm^3, flue gas velocities across the throat are about 2 m/s, and pressure drop across the throat is about 3 kPa.

In the SO_2 absorber, L/G's vary from 0.5 to 1.2 liter/Nm^3, flue gas velocities in the scrubber from 2 to 7 m/s, and pressure drops across the absorber from 1.0 to 5.0 kPa.

Flue gas velocities in a vertical demister average 6 m/s. The optimum flue gas velocity in horizontal demisters is 2.5 m/s. Increasing the velocity increases the water removal rate but also increases the entrainment of water before it can leave the demister. Pressure drop across the demister ranges from 0.2 to 0.5 kPa.

The flue gas is reheated sufficiently (75° to 95°C) to prevent condensation. Depending on the method of reheat, this requires an energy expenditure of approximately 1100 J/kg of flue gas/°C of reheat.

Regeneration tank retention times vary from 10 minutes for lime regeneration to 2 hours for limestone regeneration. A 500-MW unit with a lime regeneration system would have a regeneration tank capacity of about 100 m^3. Centrifuge or vacuum filters require a capacity of about 130 m^3/h.

3.0 CURRENT STATUS, ASSESSMENT, AND PLANS

The double alkali/gypsum process has exceeded its original design intent, minimizing scaling and maximizing availability while incorporating the major benefits of a lime/limestone system. As an additional benefit it produces gypsum of large grain size and high quality. This gypsum is more suitable for wallboard production than that produced by the direct lime/limestone method. The major disadvantages are its higher capital and operating costs compared with the direct lime/limestone system.

The double alkali system can achieve 98 percent SO_2 removal with very high lime/limestone utilization. The process operates near stoichiometric ratios and can be used effectively on high- or low-sulfur gas streams.

The success of the double alkali/gypsum system is indicated by five new systems recently installed on utility boilers. The systems were designed by Kureha-Kawasaki. Boilers range in size from 150 to 450 MW and have SO_2 inlet concentrations up to 1300 ppm.

The process is most suitable to areas with an abundance of locally available, inexpensive lime or limestone. By-

product disposal is flexible because two distinctly different methods are available. The best method is to sell the gypsum to local manufacturers of wallboard and cement. No land is required for disposal, and a sales credit reduces the cost of the system. The alternative is to use the gypsum as landfill. It can make a good construction base without any additional expense. Otherwise the system has shown excellent performance in many areas: it is applicable to both high- and low-sulfur fuel, tolerates fly ash in the scrubber unless the gypsum is to be sold, has high availability, low maintenance, and a medium to low capital cost.

Currently no new large double alkali/gypsum systems are planned in Japan for several reasons. Fuel desulfurization techniques are cost-competitive with FGD systems. The by-product market for gypsum is oversupplied. New processes for simultaneous SO_2 and NO_x removal are being developed, and many industries are holding back decisions as long as they can. Ambient SO_2 levels are declining, and the ambient standards are almost achieved in some areas.

Table 1 summarizes the status of surveyed double alkali/gypsum systems. The ranges of collection efficiencies and availability refer only to the surveyed systems. The rest of the information refers to all double alkali/gypsum systems through December 1977. Table 2 presents

Table 1. SUMMARY OF STATUS OF SURVEYED

DOUBLE ALKALI/GYPSUM FGD SYSTEMS

All Systems	
Total number of systems in operation in December 77	25
Total number of systems built	25
Total number of systems surveyed	3
Year first placed in commercial service	1973
Number of utility systems using process	7
Total MW of these facilities	1949
Number of industrial combustion systems	18
Surveyed Systems	
Range of average SO_2 removal efficiencies, %	90 - 95
Range of fly ash removal efficiency by FGD systems, %	38.4 - 80
Range of average availability factors,[a] %	95 - 98

[a] Availability factor computation see Appendix B.

Table 2. STATUS OF SURVEYED FGD INSTALLATIONS

DOUBLE ALKALI - GYPSUM PROCESS

Name	Location	Size, MW	Fuel type	Characteristics of fuel: % S	Characteristics of fuel: Heat value, 10^3 J/kg	Flue gas treated: Nm^3/s	Flue gas treated: % of total	Year FGD systems started	SO_2 removal efficiency, %	Disposition of end products	Energy required, %	Average availability, %
Sakaide Shikoku Elect. Co.	Sakaide Kagawa Prefecture Japan	650	Heavy oil C	2.0	41.36	355	100	1975	94.5	Gypsum sold for cement or wallboard	1.7[a]	95 (4/76 to 9/76)
Ichihara Power Station Showa Denko Co.	Ichihara City, Chiba Prefecture Japan	170 2 boilers	Heavy oil C	2.85	41.40	139	100	1973	90-95	Gypsum sold for wallboard	N.A.[b]	90
Tamano Heikei Salt Production Co.	Tamano City Okayama Prefecture	35 Equivalent	Heavy oil C	2.8	41.86	21	100	1976	93	N.A.	2.2[c]	99.3

[a] In addition 2.06 reheat.

[b] In addition 3.15 reheat.

[c] No reheat.

N.A. - Not available.

important characteristics of each surveyed facility, and the Appendix contains questionnaires on which more detailed information may be found.

4.0 RECENT TECHNOLOGICAL DEVELOPMENTS AND SOLUTIONS TO PROBLEM AREAS

Sodium-based double alkali systems are not subject to the scaling and erosion problems associated with direct lime/limestone-based systems because of the high solubility of sodium alkali and the sulfur-containing reaction products. The best SO_2 removal and the greatest reduction in scale formation are achieved by precise control of the pH of the scrubber solution.

4.1 CONTROL OF pH

A control point of pH 6.5 has been found to be the best for concentrated alkali system reliability and performance. Increasing the pH to about 7 increases carbon dioxide absorption and can lead to calcium carbonate precipitation and possibly to carbonate scaling. Decreasing the pH to below 6 dramatically increases the equilibrium vapor pressure of SO_2 in the gas above a concentrated active alkali solution and can lead to equilibrium-limited scrubbing conditions where outlet concentrations at or below 200 ppm are desired. Graphical evidence of this phenomenon is provided in Figure 6, in which sulfur dioxide vapor pressure

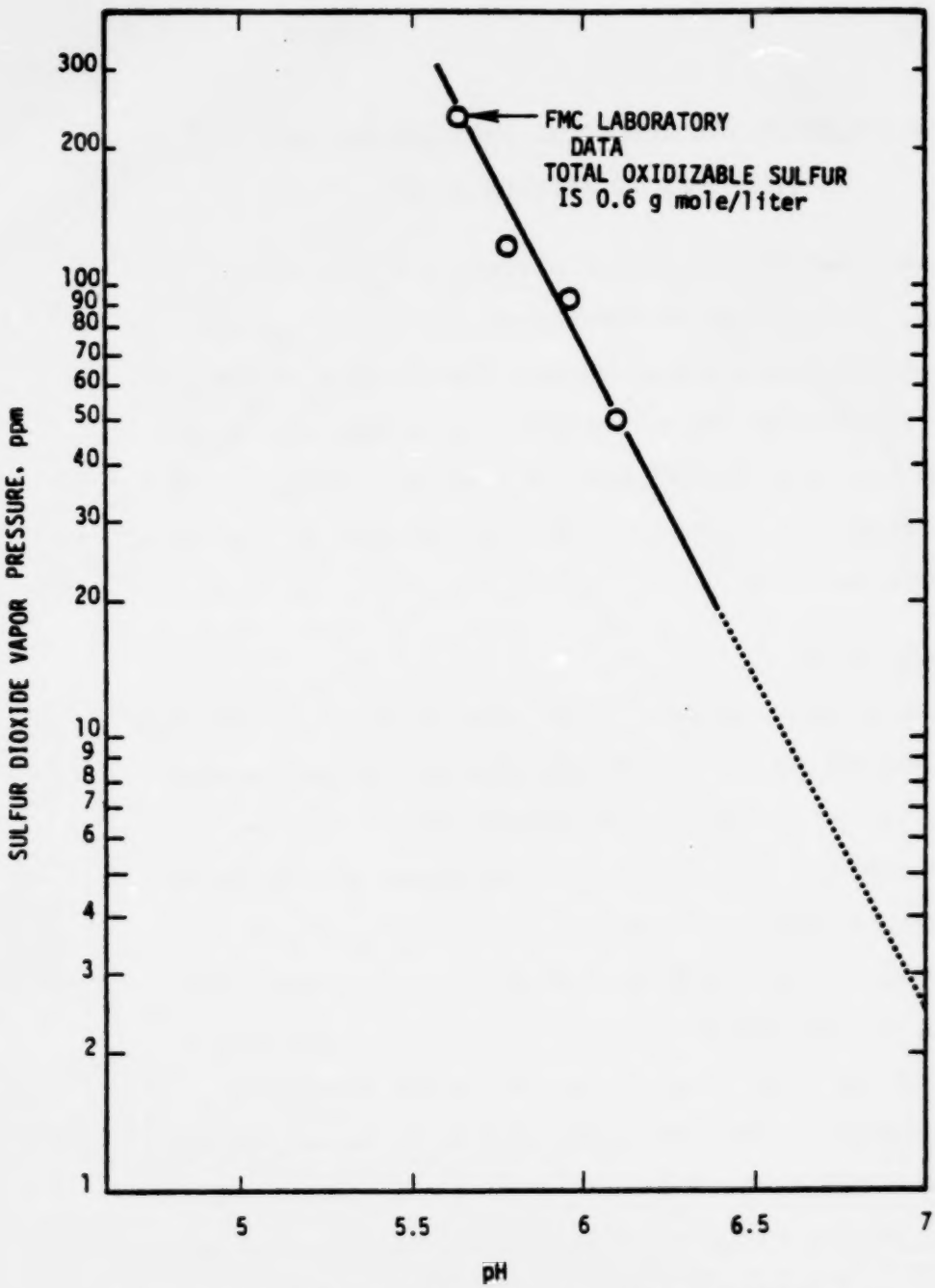

Figure 6. Concentrated active alkali systems: sulfur dioxide vapor pressure vs. pH @ 54°C

(From: the FMC Concentrated Double-Alkali Process, by Legatski L.K., K.E. Johnson, and L.Y. Lee, FMC Corp. Glen Ellyn, Illinois)

is plotted as a function of pH for a solution temperature of 54°C in a concentrated mode system. In a dilute alkali system, the SO_2 vapor pressure is much lower.

Operation within the pH range of 6 to 7 is acceptable, but a set point of pH 6.5 is preferred because the sulfite-bisulfite system is highly buffered at that pH, and a highly buffered system is able to adapt to rapidly changing boiler flue gases. In contrast, scrubbing systems operating at a pH of less than 6 can be very sensitive to and adversely affected by rapid changes in flue gas inlet concentrations.

4.2 FORMATION OF SULFATES

Sulfate formation in the scrubbing system can seriously affect system performance. Sulfates do not absorb SO_2 and are not readily regenerated into an active sodium form by reaction with lime. Sulfate also can form gypsum scale under certain conditions; thus, every attempt is made to suppress oxidation of sulfite to sulfate anywhere in the scrubber system. Oxidation is known to be a function of the rate of oxygen dissolution, scrubbing solution pH, trace contaminants (catalytic effects of fly ash or lime and soda ash impurities), ionic strength of the solution, absorber design, SO_2 and O_2 concentrations in the flue gas, and flue gas temperature. Because oxidation is influenced strongly by oxygen absorption, the number of contact stages should be

limited to two, and the regeneration reactor must be designed to minimize oxidation. In concentrated-mode systems, the unsaturated calcium sulfate mode of operation is maintained by co-precipitation of calcium sulfite and sulfate from slurry liquid saturated in calcium sulfite and unsaturated in calcium sulfate. This produces low sulfite-to-sulfate oxidation rates. Calcium sulfate can be co-precipitated with the principal sulfite precipitate up to a sulfate/sulfite mole ratio of approximately 0.2. Thus, unsaturated conditions can be maintained only at relatively low sulfite oxidation rates achievable in high SO_2/O_2 ratio emission sources. The minimum inlet gas conditions necessary to operate a concentrated-mode system reliably and economically are thus 400 to 500 ppm SO_2 and 5 percent O_2. A lower SO_2/O_2 ratio is unacceptable to the concentrated mode of operation because of the domination of the sulfate species. For flue gas applications that equal or fall below these inlet gas requirements (such as predominate in some industrial boilers, bark boilers, and paper mills), dilute-mode double alkali systems are technically feasible because they can withstand high oxidation rates (up to 100 percent).

4.3 CHLORIDE ION (Cl^-)

High chloride concentration in the scrubber system can cause problems. Water, limestone, and especially fuel can

bring chloride ion into the system. Coal-burning systems are especially susceptible to build-up of chloride ion concentrations. Chloride ion in the coal at or above 0.04 percent by weight can increase tendencies to develop scale in the scrubber loop. Chloride ion also tends to increase corrosion of internal scrubber parts, reduce the concentration of active sodium, increase the loss of sodium to the waste stream as sodium chloride (NaCl) (causing a potential water pollution problem), and reduce the sales potential of by-product gypsum. The best solution to this problem is to burn low-chloride coal. If this is not possible, a pre-scrubber with a separate liquor loop should be installed to remove hydrogen chloride (HCl) from the flue gas, and the system must be designed against chloride corrosion. Finally, a portion of the recycle liquor stream could be treated to remove chloride. Stringent water pollution control regulations could require the installation of deionizers on the wastewater purge stream. This could add substantially to the cost and require additional sodium makeup.

5.0 COST

A comparison of economics between systems in various countries is complex because economic factors differ in each country. Costs of labor, raw materials, utilities, taxes, and interest rates not only differ, but they change from year to year. Inflation increases at a different rate in each country, and consequently, the relative value of money between countries changes. A further complication results from FGD systems being built differently. Japanese systems have pH controllers, convert all the sludge to gypsum, and include spares in some systems. This complicates the comparison of two plants, even in the same country. Adjusted costs of individual installations of double alkali/gypsum FGD installations are presented in Table 3, and actual costs appear in the questionnaires in the Appendix.

Cost estimates provided by each company differ in what was included. The Shikoku Electric Company showed only capital cost and net annual cost per kWh. The Showa Denko Company presented itemized annual costs but did not itemize capital costs. Showa Denko also indicated product credits and subtracted them from annual operating costs. The Tamano

Table 3. COST DATA OF INDIVIDUAL INSTALLATIONS FOR THE DOUBLE ALKALI GYPSUM FGD SYSTEM

	Company name	Shikoku Electric Co.	Showa Denko Co.	Nakai Salt Production Co.
	Plant name	Sakaide	Ichihara Power Station	Tamano
	Location	Sakaide Kagawa Pref. Japan	Ichihara City Chiba Pref. Japan	Tamano City Okayamo Pref. Japan
1.	Reported capital costs, $[a]	40,050,000	6,664,000	1,404,000
	Year	1974	1972	1975
	Cost inflated to 1976	46,800,000	9,180,000	1,430,000
	$/kW	104	54	55
2.	Reported annual costs, $			
	Total variable costs	N.A.	3,741,000[c]	440,000[d]
	Total fixed costs[b]	8,728,200[c]	1,894,000[c]	197,000[c]
	Fixed charge factor, %	18.65	20.63	13.78
3.	Boiler operating hours	7362	8000	8000
	Capacity factor	0.642	0.88	0.82
	Generating MWh	253076	1310496	186763[e]
4.	Reported cost, mills/kWh	6.406	5.136	2.33
	Total cost, mills/kWh 4000 h		9.21	4.03
	8000 h		4.60	3.07
	Fixed cost	3.45	1.45	1.05
	Variable costs	N.A.	2.85	1.28

[a] Yen is converted to dollars for year of construction initiation and then converted to January 1976 dollars using the Chemical Engineering Plant Cost Index (1957-59 = 100, Jan 1976 = 187.1), 1 dollar = 300 yen.

[b] Fixed costs are normally based on 7 years depreciation in Japan, estimates shown are based on twenty years.

[c] Estimated, i.e. cost to build same system in U.S.

[d] Product credit or disposal costs not included.

[e] Equivalent, not an electric utility boiler.

N.A. - Not available.

plant of Nakai Salt Product Company is an aluminum sulfate/ gypsum process, slightly different from the other two systems, and therefore its costs are different.

All costs reported in the questionnaire were assumed to apply to the date of plant start-up unless otherwise indicated. (Plant start-up date and length of construction were generally reported.) The dollar cost at initiation of construction was converted to the cost of a January 1976 initiation of plant construction, using the Chemical Engineering Plant Cost Index (1957-59 = 100). Capital costs were not itemized; annual costs were. Waste disposal or by-product credits were not included.

6.0 BIBLIOGRAPHY

1. La Mantia, C.R., R.R. Lunt, R.E. Rush, T.M. Frank, and N. Kaplan. Operating Experience CEA/ADL Dual Alkali Prototype System at Gulf Power/ Southern Services, Inc. Arthur D. Little, Inc. (Presented at EPA Flue Gas Desulfurization Symposium. New Orleans, Louisiana, March 8-11, 1976.) 38 p.

2. Survey Report on SO_2 Control Systems for Non-Utility Combustion and Process Sources. PEDCo Environmental, Inc. Cincinnati, Ohio. Contract No. 68-02-2603, Task No. 4. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. May 1977.

3. Pike, D.E. Double Alkali Chemistry of Sulfur Dioxide Collection for Boiler Flue Gas Cleaning Systems. Air Pollution Industries, Englewood, New Jersey. October 1974.

4. Draemel, D. An EPA Overview of Sodium-Based Double Alkali Processes/Part I. A View of the Process Chemistry of Identifiable and Attractive Schemes. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. May 1973. p. 1-20.

5. Interess, E. Evaluation of the General Motors' Double Alkali I SO_2 Control System. Arthur D. Little, Inc. Cambridge, Massachusetts. Contract No. 68-02-1332, Task 3. Program Element No. EHE624. U.S. Environmental Protection Agency, Office of Research and Development, Washington, DC. January 1977. p. 1-93.

6. Slack, A.V., and G.A. Hollinden. Sulfur Dioxide Removal from Waste Gases. Noyes Data Corporation, Park Ridge, New Jersey. 1975. 294 p.

7. Ando, J., and G.A. Isaacs. SO_2 Abatement for Stationary Sources in Japan. Industrial Environmental Research Laboratory. Research Triangle Park, North Carolina. Contract No. 68-02-1321, Task 6, Program Element No. 1AB013. U.S. Environmental Protection Agency, Washington, DC. January 1976.

8. Isaacs, G.A. Survey of Flue Gas Desulfurization Systems. The Gulf Power Company, Southern Services Co. National Environmental Research Center. Research Triangle Park, North Carolina. U.S. Environmental Protection Agency, Washington, DC.

9. Gall, R.C., and E. J. Piasecki. The Double Alkali Wet Scrubbing System. Chemical Engineering Process, New York, 71:72-76. May 1975.

10. La Mantia, C.R., and R.R. Lunt. Dual Alkali Process for Sulfur Dioxide Removal. Chemical Engineering Process 70:66-67, New York. June 1974.

11. SO_2 Abatement for Stationary Sources in Japan. PEDCo Environmental, Inc., Cincinnati, Ohio. June 1976.

12. PEDCo Environmental, Inc. Cincinnati, Ohio. Survey Report on SO_2 Control Systems for Non-Utility Combustion and Process Sources. 68-02-2603, Task No. 4. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. May 1977.

13. The McIlvaine Company. The McIlvaine Scrubber Manual. Vol. 3, Northbrook, Illinois. 1974-1977.

14. PEDCo Environmental, Inc. Cincinnati, Ohio. Summary Report - Flue Gas Desulfurization Systems - January, February, March 1977. Division of Stationary Source Enforcement and Industrial Environmental Research Laboratory. U.S. Environmental Protection Agency, Research Triangle Park, North Carolina. 1977.

15. Zada, F.K. Status of Flue Gas Desulfurization Processes. PEDCo Environmental, Inc. Cincinnati, Ohio. Aluminum Company of America, Pittsburgh, Pennsylvania. January 13, 1975. 128 p.

16. PEDCo Environmental, Inc. Cincinnati, Ohio. Support Study for Review of New Source Performance Standards for Coal-Fired Steam Generator. Preliminary Draft. Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina. 1977.

17. Standards for Metric Practice. American Society For Testing and Materials, Philadelphia, Pennsylvania, February 1976.

APPENDIX A

APPENDIX A. QUESTIONNAIRE DATA ON THE DOUBLE ALKALI/GYPSUM FLUE GAS DESULFURIZATION PROCESS

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
1.	GENERAL DESCRIPTION			
1.1	Company	Shikoku Electric Co.	Showa Denko Co.	Naikai Salt Production Co.
1.2	Plant name	Sakaide	Ichihara Power Station	Tomano
1.3	Preparer's name and mailing address	J. Ando 1-13-27 Kasuga, Bunkyo-Ku, Tokyo	Jumpei Ando 1-13-27 Kasuga, Bunkyo-Ku, Tokyo	Jumpei Ando 1-13-27 Kasuga, Bunkyo-Ku, Tokyo
1.4	Principal plant products	Electrical Power	Electrical Power	Salt
	Brief description of area surrounding plant site and factors of choice of FGD system	Located on reclaimed land at Seto inland sea. SO_2 emission is kept below 610 Nm^3/h by using a high efficiency FGD on the 450 MW boiler and medium sulfur oil without FGD on others.	Located near Ichihara and Chiba cities. Two boilers use sodium limestone gypsum process, two use limestone gypsum process and two burn low sulfur oil to control emissions of SO_2.	Located at Mizushima Industrial Complex district facing the Seto inland sea. FGD needed because of stringent SO_2 regulations.
1.5	Plant location	Sakaide, Kagawa Prefecture	Ichihara City, Chiba Prefecture	Tomano City, Okyama Prefecture
1.6	STATION CAPACITY			
1.6.1	Total boiler generating capacity MW/number of boilers	1300/4	620/6	26
1.6.2	Capacity served by FGD system, MW	450	170	26
	% of boiler flue gas treated by FGD system	100	100	100
	Number of boilers served	1	2	1

Unless otherwise noted Nm^3/h = m^3/h (0°C and 101.325 kPa)

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
1.7	TYPE OF FGD SYSTEM	Sodium limestone/gypsum	Sodium limestone/gypsum	Aluminum sulfate/gypsum
1.8	Number of scrubber trains in parallel	2 (no standby)	4 (no standby)	1
1.9	FGD system installed on new or existing boiler	Existing	Existing	Existing
1.10	STATUS OF FGD SYSTEM operational or under construction date started	Operational/October 1975	Operational/May 1973	Operational/April 1976
2.	BOILER DATA			
2.1	Boiler identification number	Sakaide No. 1	No. 1, No. 2	N.A.
2.2	Maximum continuous heat input, MW	1084	256 (per boiler)	N.A.
2.3	Max. cont. net generating capacity, MW	450	85 (per boiler)	26
2.4	Flue gas flow rate, Nm^3/h temperature at induced draft fan, °C	1,278,000 135	500,000 (2 boilers) N.A.	82,000 (forced draft fan) N.A.
2.5	Boiler manufacturer	Mitsubishi heavy industries	N.A.	N.A.
2.5.1	Fuel type	Heavy oil and crude oil	Heavy oil	Heavy oil C
2.5.2	Firing configuration	Corner firing	Front wall firing	Supersonic burner
2.6	Year placed in service	1973	1961	1972
2.7	Type service	Base load	Base load	Base load
2.8	Stack height, m	200	120	45
2.8.1	Flue gas exit velocity, m/s	37.4	20	16
2.9	Average boiler operation, h/yr	7362/1976	8000/1976	8000/1976
2.10	Average boiler capacity factor	0.842	0.88	0.82

N.A. - Not available.

(continued)

APPENDIX A (continued)

	KUBOTA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	OKUA MINING CO. Tomano
2.11 COAL COMPOSITION	None	None	None
2.11.1 Lower heating value, kg/kg kcal/kg			
2.12.2 Sulfur content, %			
2.11.3 Ash content, %			
2.11.4 Chloride content, %			
2.11.5 Moisture, %			
2.12 OIL COMPOSITION			
2.12.1 Grade/heating value, kg/kg kcal/kg	Heavy oil C/10,420 kcal/kg	Heavy oil C/10,500 kcal/kg	Heavy oil C/9,700 kcal/kg
2.12.2 Sulfur content, %	2.0	2.8-2.9	2.8
2.12.3 Ash content, %	Below 0.1	0.03	0.014
2.12.4 Chloride content, %	N.A.	N.A.	N.A.
2.13 PARTICULATE REMOVAL EQUIPMENT NOT PART OF FGD SYSTEM			
2.13.1 Type	Electrostatic precipitator	None	None
2.13.2 Manufacturer	Mitsubishi Heavy Industries		
2.13.3 Efficiency, % removal	over 80 (design)		
2.13.4 Emission rate, mg/Nm^3	37		

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
3.	EMISSION REGULATIONS			
3.1	ATMOSPHERIC EMISSIONS			
3.1.1	SO_2	630 Nm^3/h (4 boilers)	120 ppm	200 ppm
3.1.2	Particulate	50 mg/Nm^3	50 mg/Nm^3	N.A.
3.1.3	NO_x	230 ppm	180 ppm	N.A.
3.1.4	Other	N.A.	N.A.	N.A.
3.1.5	Legal basis	Agreement with Kagawa Prefecture, except NO_x is national standard	Regulation by Chiba Prefecture	Agreement with local government
3.2	RESTRICTIONS APPLICABLE TO LIQUID OR SOLID WASTES LEGAL BASIS			
4.	FGD SYSTEM INFORMATION			
4.1	Process name	Sodium limestone/gypsum	Showa Denko Sodium Limestone gypsum process	Aluminum sulfate limestone Gypsum process
4.1.1	Supplier or design company	Kureha Chem. and Kawasaki Heavy Industries	Showa Denko	Dowa Mining Co.
4.1.2	WHY WAS THIS SYSTEM CHOSEN?	High SO_2 removal efficiency	Stable operation and high SO_2 removal efficiency	Stable reliable operation, low cost
4.1.3	PROCESS DESCRIPTION	See figures A and B	See figure C	See figure D

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
4.2	SO_2 SCRUBBING AGENT			
4.2.1	Name	Sodium hydroxide	Sodium hydroxide (sodium sulfite)	Aluminum sulfate Soln.
4.2.2	Chemical composition	N.A.	N.A.	Al SO_4 H_2O
	Components/% by weight	N.A.	N.A.	Al/4.2 SO_4/22.6 H_2O/73.2
4.2.3	Stoichiometric requirement	N.A.	N.A.	N.A.
4.2.4	Addition rate, kg/h	N.A.	N.A.	8 ℓ/h (for makeup)
4.2.5	Particle size, mm	N.A.	N.A.	N.A.
4.3	REGENERATION AGENT			
4.3.1	Name	Limestone	Limestone	Limestone
4.3.2	Composition	$CaCO_3$/over 98 MgO/below 0.3	$CaCO_3$/over 96	$CaCO_3$/99 (approx) MgO/0.5 SiO_2/0.5
4.3.3	Stoichiometric requirement	1.0	1.0	1.0
4.3.4	Addition rate, kg/h	8182		400
4.3.5	Particle size, mm	N.A.	Below 0.102	80% below 0.063
4.4	SALABLE PRODUCT			
4.4.1	Name	Gypsum	Gypsum	Gypsum
4.4.2	Chemical composition Components/% by weight	$CaSO_4 \cdot 2H_2O$/90 H_2O/10	$CaSO_4 \cdot 2H_2O$ Moisture/below 10	$CaSO_4 \cdot 2H_2O$ Moisture/below 10
4.4.3	Amount generated/year, tons	90,000	64,000	6000
4.4.4	Sales potential	Cement and wallboard	Sold for wallboard	N.A.

(continued)

APPENDIX A (continued)

		BURIBA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
4.4.5	BY-PRODUCT RECOVERY PROCESS	N.A.	N.A.	N.A.
	Name			
	Vendor			
	Process description			
4.4a	Intermediate product composition			
4.5	BY-PRODUCT NUMBER 2	N.A.	N.A.	N.A.
4.6	WASTE STREAMS			
4.6.1	Type of waste	None	Wastewater	Wastewater
	I water	N.A.	N.A.	N.A.
4.6.2	Amount generated in 1 hour at full load, m^3/h	N.A.	1.6 - 5.4	0.03
4.6.3	Disposal method	N.A.	N.A.	N.A.
4.6.4	Disposal site and distance from plant, km	N.A.	N.A.	N.A.
4.6.5	Potential effects on environment from sludge disposal	N.A.	N.A.	N.A.
4.7	SLUDGE FIXATION PROCESS	None	None	None

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
4.8	FGD EQUIPMENT DATA			
4.8.1	FIRST SCRUBBER			
4.8.1.1	Type	Packed tower	None	Spray
4.8.1.2	Liquid(solid)/gas ratio, t/Nm^3	1.2		2.5
4.8.1.3	Gas velocity through absorber or scrubber, m/s	2.0		5.0
4.8.1.4	Absorbend (absorbent) flow, m^3/h	N.A.		N.A.
4.8.1.5	Gas flow entering scrubber, Nm^3/h	1,278,000		82,000
4.8.1.6	Gas temperature °C	135		170
	SO_2 in • out, ppm	1270 • 69		1500 • 1429
	% removal	94.5		4.7
	Particulate in • out, mg/Nm^3	17.7 • 10.9		20 • 20
	% removal	38.4		
	NO_x in • out, ppm	181 • 184		300 • 300
	CO_2 in • out, ppm	13.5 • 13.4		10.5 • 10.5
	Other in • out	N.A.		N.A.
4.8.1.8	Material of contruction	Mostly SS41		Carbon steel
4.8.1.9	Type of lining	Rubber		Acid brick
4.8.1.10	Absorbent and sand losses	N.A.		N.A.
4.8.1.11	Number of stages	2 (upper) (lower)		1
4.8 1.12	Packing type and composition	Plastic tellerette (upper) Plastic grid (lower)		None

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
4.8.2	SECOND SCRUBBER	N.A.		
4.8.2.1	Type		Vertical cone	Packed tower
4.8.2.2	Liquid (solid)/gas ratio, ℓ/Nm^3		0.5-1.0	10
4.8.2.3	Gas velocity through absorber or scrubber, m/s		4-7	0.9
4.8.2.4	Absorbent (absorbent) flow, m^3/h		N.A.	N.A.
4.8.2.5	Gas flow entering scrubber, Nm^3/h		130,000 (per scrubber)	82,000
4.8.2.6	Gas temperature, °C		140	56
4.8.2.7	Gas composition			
	SO_2 in • out, ppm		1300-1500 • 75-100	1429 • 100
	% removal		90-95	93
	Particulate in • out, mg/Nm^3		100-200 • Below 50	20 • 10
	% removal		75-80	50
	NO_x in • out, ppm		N.A.	300
	% removal		N.A.	N.A.
	CO_2		N.A.	10.5%
	Other			N.A.
4.8.2.8	Material of construction		SUS 304 and 316	Carbon steel
4.8.2.9	Type of lining		Rubber and FRP	Flake lining
4.8.2.10	Number of stages		1	1
4.8.2.11	Packing type and composition		None	Plastic

(continued)

APPENDIX A (continued)

		KUBOTA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
4.8.3	CLEAR WATER TRAY			
4.8.3.1	Type	Water spray	None	None
4.8.2.3	L/G ratio, ℓ/Nm^3	0.01		
4.8.3.3	Source of water	Industrial water		
4.8.4	DEMISTER			
4.8.4.1	Type	Torellette	Special impingement plate	Wire mesh
4.8.4.2	Number of passes (stages)	1	3	3
4.8.4.3	Material of construction	Polyethylene	N.A.	SUS 316
4.8.4.4	Method of cleaning	Water wash	Water spray	Water wash
4.8.4.5	Flow rate during cleanings, Nm^3/h	80	N.A.	N.A.
4.8.4.6	Frequency and duration of routine cleaning	Once weekly for 8-9 hours	N.A.	6 times daily
4.8.5	REHEATER			
4.8.5.1	Type	After burner	After burner	None
4.8.5.2	Heat load, kcal/h	23,000,000	6,700,000	
4.8.5.3	Heat transfer surface area, m^2	None	None	
4.8.5.4	Temperature of gas in, °C out, °C	80 135	55 130-140	
4.8.5.5	Reheater materials of construction	SS 41 (lined with refractories)	N.A.	

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
4.8.6	SCRUBBER TRAIN PRESSURE DROP, mbar			
4.8.6.1	First scrubber	11		1-2
4.8.6.2	Second scrubber		29-69	7-8
4.8.6.3	Clear water tray	1		
4.8.6.4	Demister	2	1-5	2-4
4.8.6.5	Reheater			
4.8.6.6	Ductwork	14	15	1-2
4.9.6.7	Total FGD system	30	19-69	17
4.8.6.8	I.D. fan	35	69	17 (forced)
4.9	SYSTEM MAKEUP WATER ADDED, m^3/h	12.5	13.6	4.1
4.9.1	Water added per mole of SO_2 removed, m^3/kg mol	0.2 t/mol	0.9 kg/mol	
4.9.2	Source of makeup water	Industrial water	River water	Water from boiler
4.9.3	Where is water added	Absorber demister	Scrubber and gypsum wash	Demister and gypsum centrifuge
4.10	BYPASS SYSTEM			
4.10.1	Can flue gas be bypassed?	Possible	Possible	N.A.
4.10.2	Height of gas discharge when bypassed, m	N.A.	120	N.A.
4.11	SLURRY DATA,	N.A.		N.A.
	pH/% solids by weight			
4.11.1	Fresh slurry to scrubber		3.5/10-30	
4.11.2	Recirculating slurry		N.A.	
4.11.3	Waste slurry		N.A.	

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
4.12	SPENT SLURRY SYSTEM	N.A.	N.A.	N.A.
4.12.1	Clarifiers			
4.12.1.1	Number			1
4.12.1.2	Diameter, m			6.5
4.12.2	Solid separation			N.A.
4.12.2.1	Type			Centrifuge
4.12.2.2	Cloth area, m^2			2.6
4.12.2.3	Capacity (wet cake), kg/h			400
4.13	SPACE REQUIRED, EXCLUSIVE OF SLUDGE DISPOSAL AREA, m^2	12,000 SEE FIGURE A	5000	84
4.14	PERSONNEL, man-hours/year			
4.14.1	Operating	N.A.	N.A.	N.A.
4.14.1.1	Specialists	33,000	1000 (1 person - shift)	1000
4.14.1.2	Unskilled	16,000	17,500 (2 persons - 4 shifts)	8000
4.14.2	Maintenance and repair	109,900	1000 (1 person - 1 shift)	1000
4.15	ENVIRONMENTAL LOAD			
4.15.1	Waste water treatment process	N.A.	N.A.	Neutralization
4.15.2	Concentrations	N.A.	N.A.	Cl/250 ppm Hg/10g/t
4.15.3	Appearance of flue gas discharge-any mist problems?	Slight plume in winter, no mist problems	Clear discharge-no mist problems	Plume soon disappears, no mist problems

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	IKNA MINING CO. Tomano
4.16	FGD SYSTEM OPERATION			
4.16.1	% of boiler load FGD can operate	Full range	50-100[a]	10-100
4.16.2	Lag time	9 MW/min	None	1 minute
5.	ACTUAL COSTS FOR TOTAL FGD INSTALLATION			
	year	1974		1976
5.1	CAPITAL COSTS, $	N.A.	N.A.	N.A.
5.1.1	Scrubber			
5.1.2	Absorbent/regeneration system			
5.1.3	Sludge disposal system			
5.1.4	Site improvements			
5.1.5	Land, roads, tracks, substation			
5.1.6	Engineering costs			
5.1.7	Contractor's fee			
5.1.8	Interest during construction			
	Construction period, months	18	5	8
5.1.9	License fees			
5.1.10	Other			
5.1.11	Total, $	40 million (¥ 12 billion approx)	6.7 million (¥ 2 billion)	1.4 million (¥ 420 million)
	$/kW			
		¥ - Yen	¥ - Yen	¥ - Yen

[a] Below 50% load only one or two scrubbers can be used.

(continued)

APPENDIX A (continued)

		KUBOTA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
5.2	ANNUAL COSTS, $	N.A.		
5.2.1	Interest		380,000 (¥ 114 million)	42,000 (¥ 12.6 million)
5.2.2	Depreciation		851,333 (¥ 256 million)	140,000 (¥ 42 million)
5.2.3	Insurance and taxes		160,000 (¥ 48 million)	10,000 (¥ 1.0 million)
5.2.4	Total fixed costs		1,393,333 (¥ 418 million)	192,000 (¥ 57.6 million)
5.2.5	Labor and overhead Average man hour cost cost/year		100,000 (¥ 30 million)	48,000 (¥ 14.4 million)
5.2.6	Raw material breakdown Unit cost Annual cost		See A	68,000 (¥ 20.4 million)
5.2.7	Utilities, unit cost, annual costs		Unit Annual	Unit Annual
	Electricity		See B 576,000(¥172.8 million)	See C 160,000(¥48.0 million)
	Water		27,667(¥8.3 million)	9,169 (¥ 2.75 million)
	Steam		29,000(¥8.7 million)	0
	Fuel		1.15million(¥345.6million)	0
5.2.8	Maintenance cost per year		200,000 (¥ 866.2 million)	42,000 (¥ 12.6 million)
5.2.9	Other costs, list		None	42,000 (¥ 12.6 million)
5.2.10	Total variable cost per year		2.89 million (¥ 866.2 million)	279,000 (¥ 83.7 million)
5.2.11	Total annual cost	14.3 million (¥ 4.3 billion approx. 7 years depreciation)	4.12 million (¥ 1236 million) 7 years depreciation	621,000 (¥ 186.3 million) 7 years depreciation
5.3	WASTE DISPOSAL COST, $ Unit cost Annual cost	N.A.	N.A.	N.A.
5.4	BY-PRODUCT CREDIT, $	N.A.	N.A.	N.A.
5.4.1	By-product 1		Annual-136,667 (¥ 41 million)	
5.4.2	By-product 2			

(continued)

APPENDIX A (continued)

		KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tomano
5.5	NET ANNUAL COST, mills/kWh	N.A.		
5.5.1	Cost last year		3.77 (¥ 1.13)	2.33 (¥ 0.70)
5.5.2	Cost at maximum continuous rating		3.57 (¥ 1.07)	N.A.
5.5.3	Cost withouth FGD		N.A.	N.A.
6.	ENERGY REQUIREMENTS, kWh		N.A.	
	Pumps	N.A.		150
	Fan	N.A.		800
	Reheat	N.A.		0
	Reductant chemicals	N.A.		N.A.
	Other	N.A.		140
	Total	7900 excluding reheat		590

(continued)

APPENDIX A (continued)

	KUREHA CHEM. AND KAWASAKI HEAVY INDUSTRIES Sakaide	SHOWA DENKO Ichihara Power Station	DOWA MINING CO. Tamano
7. PERFORMANCE MISSING, OPERATING PROBLEMS AND SOLUTIONS			
7.1 Scrubber system	A net was installed among packing to prevent blow-ups	N.A.	None
7.2 Demister	N.A.	Piping of the drain system had to be modified to stop plugging.	None
7.3 Reheater	N.A.	N.A.	None
7.5 Sludge system	N.A.	N.A.	None
7.6 Regeneration system	N.A.	Piping had to be modified to stop plugging.	Portion of lining eroded but has been repaired with no further problems.
7.7 By-product system	Unusual vibration of centrifuge solved by uniform charge of slurry and control of concentration.	N.A.	None
7.8 Other	N.A.	N.A.	None
7.9 Availability	95% (April-September 1976)	98%	99.1

Appendix A (Continued).

A RAW MATERIAL BREAKDOWN

	Unit Cost	Annual Cost
NaOH -	$76.67/ton (¥ 23,000/ton)	$265,000 (¥ 79.5 million)
H_2SO_4 -	$33.33/ton (¥ 10,000/ton)	$192,000 (¥ 57.6 million)
$CaCO_3$ -	$15.00/ton (¥ 4,500/ton)	$345,667 (¥ 103.7 million)

B UTILITIES

	Unit Cost
Electricity -	$19.67/mwh (¥ 5,900/mWh)
Water -	$0.06/ton (¥ 18/ton)
Steam -	$6.33/ton (¥ 900/ton)
Fuel -	$100/kl (¥ 30,000/kl)

C UTILITIES

	Unit Cost
Electricity -	$0.04/Kwh (¥ 12/kWh)
Water -	$0.20/ton (¥ 60/ton)

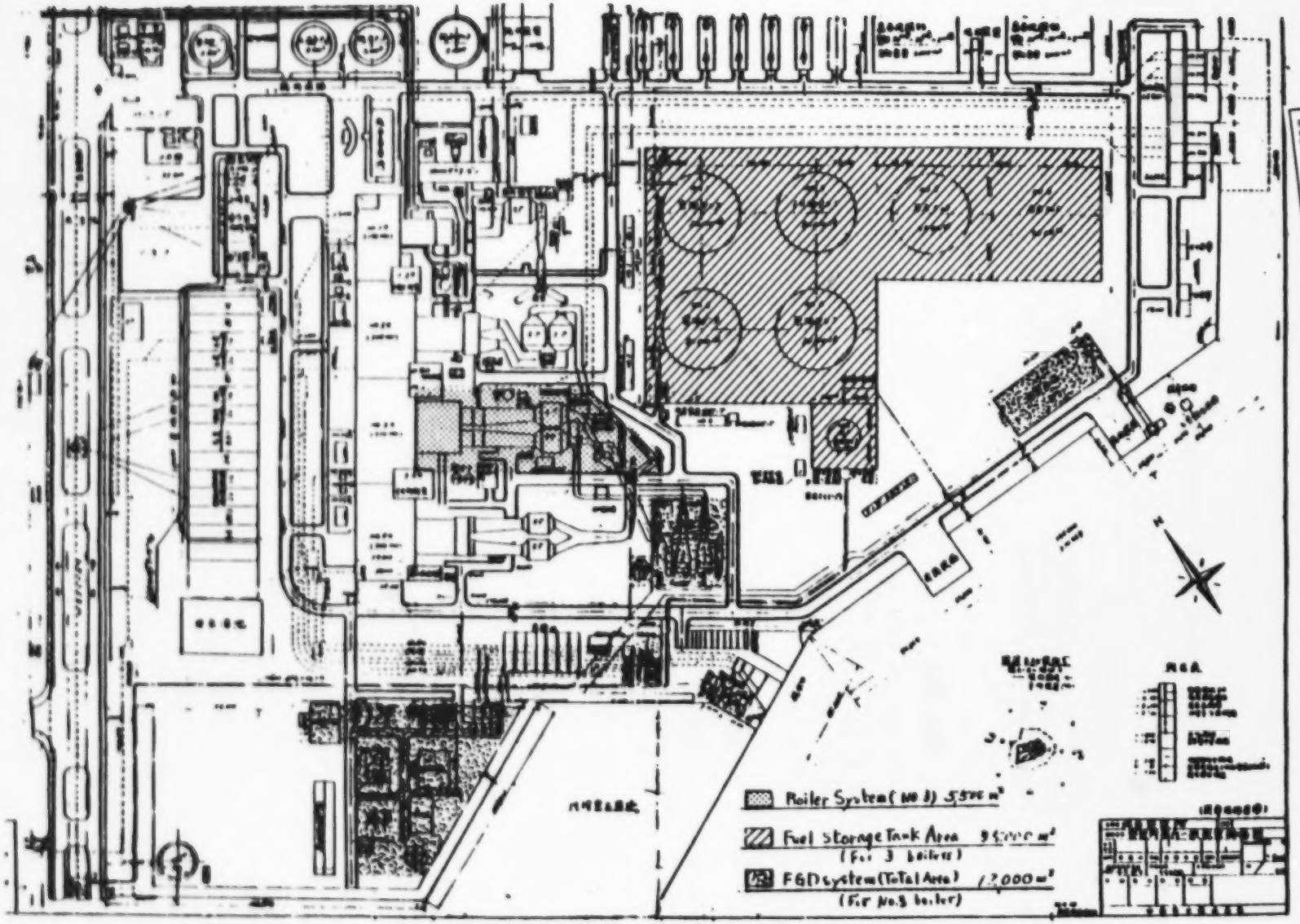

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Figure A. Shikoku Electric Co.

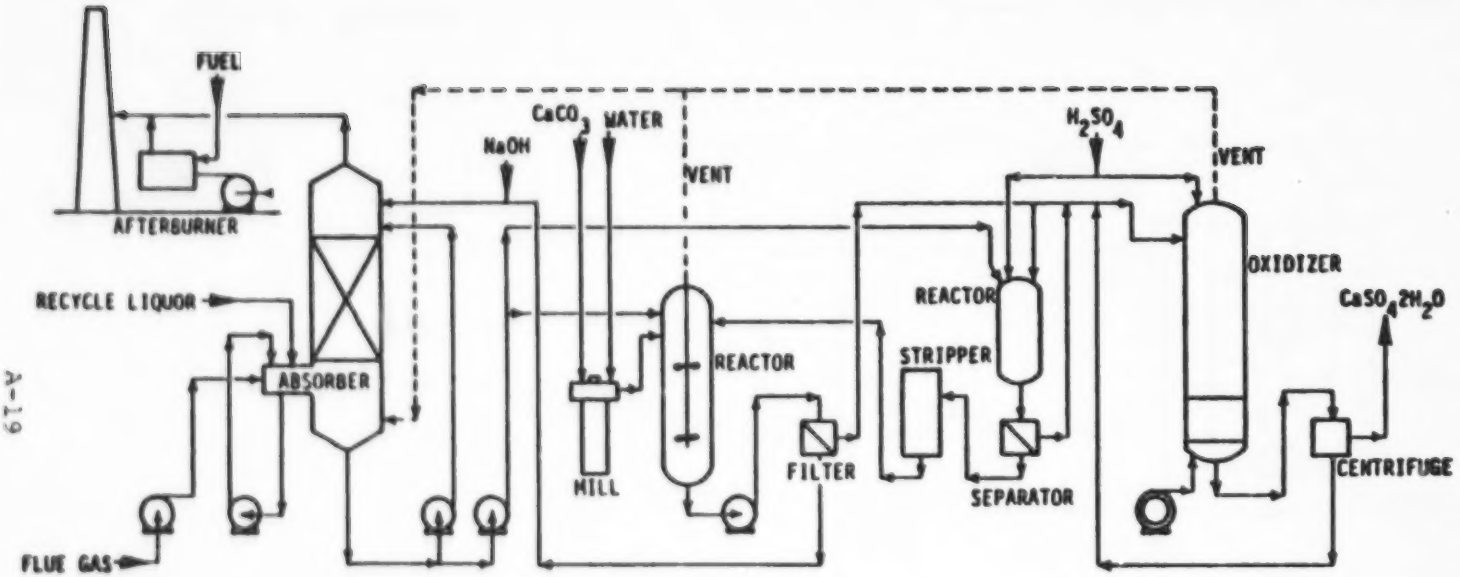


Figure B. Flowsheet of Kureha-Kawasaki process at Shikoku Electric Co.

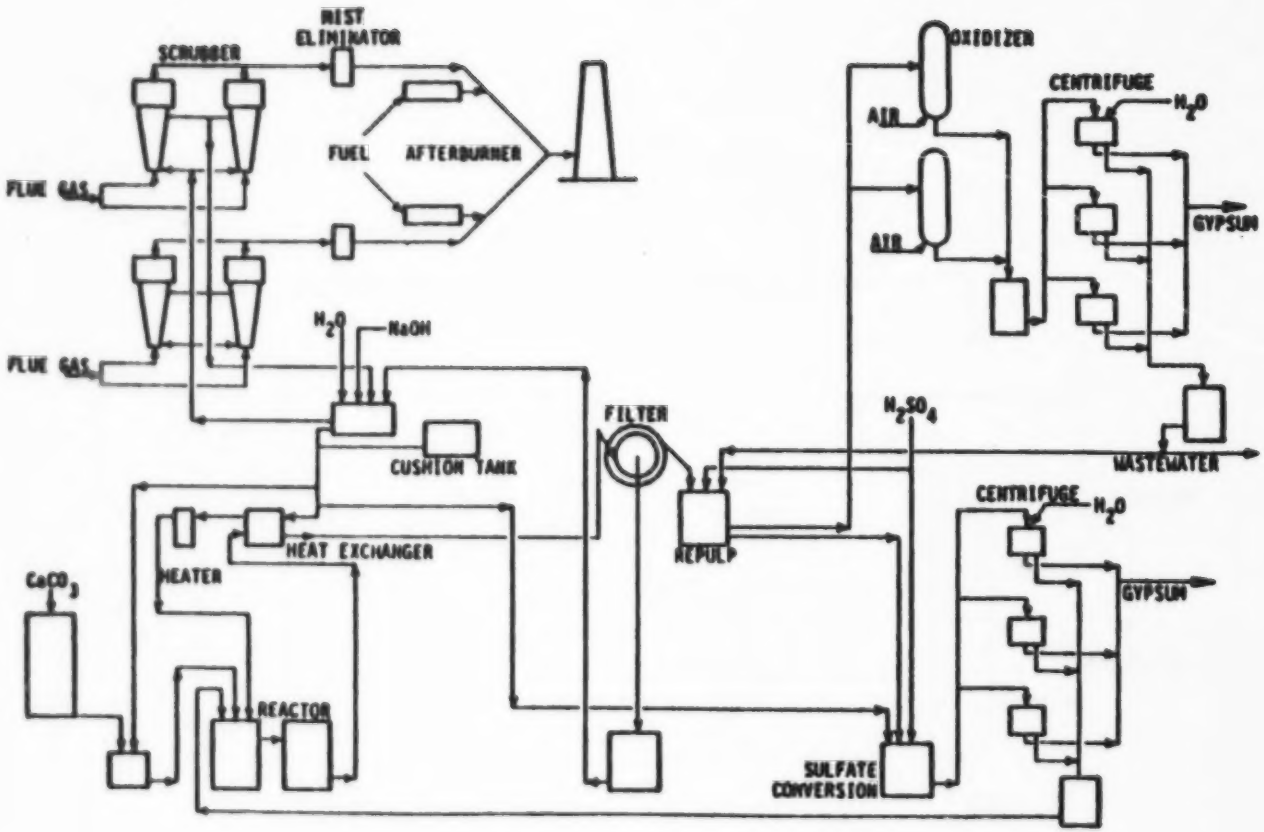
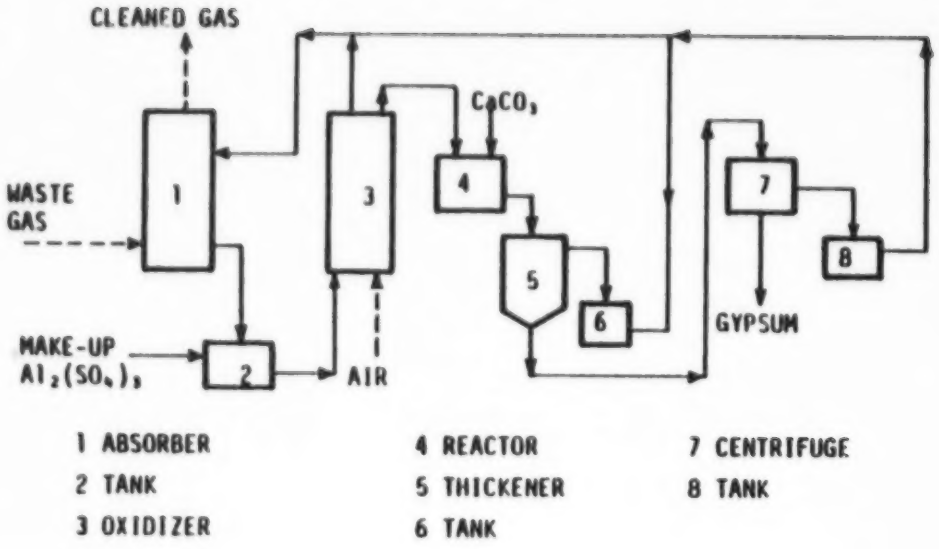


Figure C. Flowsheet of sodium limestone gypsum process (Ichihara Power Station).



1 ABSORBER
2 TANK
3 OXIDIZER
4 REACTOR
5 THICKENER
6 TANK
7 CENTRIFUGE
8 TANK

Figure D. Flowsheet of Dowa aluminum sulfate - limestone process.

APPENDIX B

APPENDIX B

Availability Factor Computation

1. Divide boiler capacity by the number of modules and obtain MW/module = x

2. Multiply boiler capacity by number of hours during selected period (MW) = a

3. Add all down times due to module trouble for all modules during selected period = b

4. Add all down times due to boiler trouble or reduction in electricity demand for all modules during period = c

5. Availability factor = $\frac{[a - x\ (b + c)]\ 100}{a - x\ c}$ = %

APPENDIX C

APPENDIX C. CONVERSION FACTORS USED IN THIS STUDY

Multiply		By	To Obtain
ft	x	0.30480	m
ft^3	x	28.317	liter
Btu/lb	x	0.55554	kcal/kg
	x	2.32593	kJ/kg
Btu/h	x	$0.29300 \cdot 10^{-3}$	kW
acfm	x	1.6992	m^3/h
$lb/10^6$ Btu	x	1.80005	mg/kcal
	x	0.42993	mg/kJ
gal/min	x	3.78541	liter/min
	x	0.22712	m^3/h
gal/1000 ft^3	x	0.13368	liter/m^3
ft^3/min	x	1.6992	m^3/h
kcal	x	4.1868	kJ
kcal	x	$1.163 \cdot 10^{-3}$	kWh
scfm	x	1.60776	Nm^3/h
gr/ft^3	x	2.288	g/m^3
gr/scfm	x	2.4182	g/Nm^3
in. H_2O	x	0.2488	kPa
in. H_2O	x	2.54	mbar
gr	x	0.0648	g
ton (short)	x	1.9072	metric ton

1 MW corresponds to 3200 to 3500 m^3/h flue gas (coal)

1 scfm = 28.317 l (60°F / 14.7 psia)
(15.5°C / 760.28 torr)

26.796 l (0°C/760 torr) = 0.026796 Nm^3

6 SO_2 = 2.927 g/liter

MW = electric power generated unless other wise specified